PROCESS DESIGN OF TURBOEXPANDER BASED NITROGEN LIQUEFIER

A THESIS SUBMITTED IN PARTIAL FULFILLMENT OF THE REQUIREMENTS FOR THE DEGREE OF

Master of Technology
in
Mechanical Engineering

By

Balaji Kumar Choudhury

Department of Mechanical Engineering
National Institute of Technology
Rourkela
2009

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Under the guidance of
Prof. Ranjit Kumar Sahoo

Department of Mechanical Engineering
National Institute of Technology
Rourkela
2009

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Rourkela

CERTIFICATE

This is to certify that the thesis entitled, **"PROCESS DESIGN OF TURBOEXPANDER BASED NITROGEN LIQUEFIER"** submitted by **Mr. Balaji Kumar Choudhury** in partial fulfillment of the requirements for the award of Master of Technology Degree in **Mechanical Engineering** with specialization in **Thermal Engineering** at the National Institute of Technology, Rourkela (Deemed University) is an authentic work carried out by him under my supervision and guidance.

To the best of my knowledge, the matter embodied in the thesis has not been submitted to any other University/ Institute for the award of any degree or diploma.

Date:

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I am extremely fortunate to be involved in an exciting and challenging research project like "process design of turboexpander based nitrogen liquefier". It has enriched my life, giving me an opportunity to work in a new environment of Aspen Hysys. This project increased my thinking and understanding capability as I started the project from scratch.

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Abstract

Our country is still dependent on imports for most of its needs in cryogenic refrigerators and liquefiers. These products are proprietary in nature which makes it very expensive for its cost and maintenance. Also as a strategy of promoting the sale, the company changes their model by limiting the spare parts of the earlier model to enforce on the customers to buy the new product. So it is imperative that our country develops an indigenous nitrogen liquefier to meet the need of liquid nitrogen. With support from the Department of Atomic Energy, our institute has initiated a program on development and study of a nitrogen liquefier of intermediate capacity in the range of 10-50 litres/hr by using technologies already developed in our country.

The process is based on a suitable modified Claude cycle which minimizes the number of heat exchangers and also takes care to accommodate the in house developed turbo expander. The process design is carried out using the standard calculation procedure and is validated by using process simulation software, Aspen Hysys.

Parametric analysis is carried out to access the role of different component efficiencies in predicting overall system efficiency at the design and off design conditions. In this analysis, the available turbo expander efficiency is considered to evaluate the feasible heat exchanger efficiency in order to optimize the plant efficiency. The thermodynamic parameters (temperature, pressure, pinch point temperature) are evaluated to obtain the optimum mass fraction through turbo expander for maximum liquid yield. This investigation not only gives the analysis of nitrogen liquefier, but also it will act as a basic frame work for any liquefier and helium liquefier in particular as a future mission.

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Nomenclature

HX1 = Heat exchanger 1

HX 2a = Heat exchanger 2 (non-condensing part)

HX 2b= Heat exchanger 2(condensation part)

Tex = Turboexpander

T = Temperature

P=Pressure

h = specific enthalpy

s= specific entropy

x= dryness fraction

1= Compressor inlet

2= Compressor outlet & HX 1 inlet (hot inlet, H.P side)

3 = hot outlet of HX 1 & Turboexpander inlet

4 = hot outlet of HX 2 & J.T Valve inlet

p= Pinch point in low pressure stream, inside HX2

5 = J.T Valve out & Phase separator in

6 = Turboexpander exit (actual state)

6' = Turboexpander exit (isentropic state)

7 = Exit of mixer 2 & cold inlet for HX 2

8 = cold outlet of HX 2 & cold in for HX 1

9 = cold out of HX 1 & mixer 1 inlet

f, g indicate saturated liquid and saturated vapor states respectively

η = efficiency of turboexpander

α = mass fraction of nitrogen diverted through turboexpander, mt/m

$\varepsilon 1$ = Effectiveness of HX 1

$\varepsilon 2$ = Effectiveness of HX 2

y = yield, liquid nitrogen separated per mass of gas compressed in the separator, mf/m

mf = mass of liquid nitrogen separated in the phase separator

m=mass of nitrogen delivered from compressor, kg/sec

mt=mass of nitrogen gas diverted through turboexpander, kg/s

Chapter 1

Introduction

1. INTRODUCTION

1.1 Principle of Liquefaction

Liquefaction of gases is always accomplished by refrigerating the gas to some temperature below its critical temperature so that liquid can be formed at some suitable pressure below the critical pressure. Thus gas liquefaction is a special case of gas refrigeration and cannot be separated from it. In both cases, the gas is first compressed to an elevated pressure in an ambient temperature compressor. This high-pressure gas is passed through a countercurrent recuperative heat exchanger to a throttling valve or expansion engine. Upon expanding to the lower pressure, cooling may take place, and some liquid may be formed. The cool, low-pressure gas returns to the compressor inlet to repeat the cycle. The purpose of the countercurrent heat exchanger is to warm the low-pressure gas prior to recompression and simultaneously to cool the high-pressure gas to the lowest temperature possible prior to expansion. Both refrigerators and liquefiers operate on this basic principle.

In a continuous refrigeration process, there is no accumulation of refrigerant in any part of the system. This contrasts with a gas liquefying system, where liquid accumulates and is withdrawn. Thus, in a liquefying system, the total mass of gas that is warmed in the countercurrent heat exchanger is less than that of the gas to be cooled by the amount liquefied, creating an imbalance mass flow in the heat exchanger. In a refrigerator the warm and cool gas flows are equal in the heat exchanger. This results in what is usually referred to as a "balanced flow condition" in a refrigerator heat exchanger. The thermodynamic principles of refrigeration and liquefaction are identical. However the analysis and design of the two systems are quite different because of the condition of balanced flow in the refrigerator and unbalanced flow in liquefier systems.

The Joule-Thomson coefficient is a property of each specific gas. It is a function of temperature and pressure, and may he positive, negative, or zero. For instance, hydrogen, helium, and neon have negative J-T coefficients at ambient temperature. Consequently, to be used as refrigerants in a throttling process they must first be cooled either by a separate pre coolant liquid. Only then will throttling cause a further cooling rather than a heating of these gases.

Another method of producing low temperatures is the adiabatic expansion of the gas through a work-producing device such as an expansion engine. In the ideal case, the expansion would be reversible and adiabatic and therefore isentropic. In this case, we can

define the isentropic expansion coefficient which expresses the temperature change due to a pressure change at constant entropy. An isentropic expansion through an expander always results in a temperature decrease. Whereas an expansion through an expansion valve may or may not result in a temperature decrease. The isentropic expansion process removes energy from the gas in the form of external work, so this method of low-temperature production is sometimes called the external work method.

1.2 Requirement of nitrogen liquefier

Nitrogen liquefier used to produce liquid nitrogen. Because of its low production cost and relatively higher levels of safety is the most common cooling medium in the cryogenic temperature range above 77 K. The application covers such diverse areas as:

- Pre coolant in production of liquid helium and low temperature refrigerators
- Cryotreatment of critical metallic components such as hubs, milling cutters, knives, rollers, needles, dies and punches, bearings and precision measuring equipment,
- Preservation of live biological material as blood, animal and human sperms, embryos, bacterial cultures etc
- Cold trap in vacuum systems and in adsorption pumps, and
- Miscellaneous laboratory and industrial applications.

1.3 Production of Liquid Nitrogen

In some parts of our country, it is possible to buy liquid nitrogen from bulk suppliers at low cost. But in most cases, including some major metropolitan areas, a laboratory needs to operate its own liquid nitrogen generator. There are three major international suppliers of nitrogen liquefiers in our country:

- Stirling Cryogenics of Netherlands,
- Linde AG, Germany, and
- Consolidated Pacific Industries, USA.

The liquefier from Stirling Cryogenics is based on the integral Philips-Stirling Cycle, while the latter two use turbine for cold production. The Linde turbine uses gas bearings, while the CPI machine uses antifriction bearings. The plants are enormously expensive to buy and to maintain and owners are often forced to buy new plants due to non-availability of proprietary spares. It is imperative that our country develops an indigenous nitrogen liquefier of capacity in the range 10 to 50 l/hour. Hence a nitrogen liquefier is to be designed.

1.4 Objectives of the work

Prior to the making of the turboexpander based nitrogen liquefier, the thermodynamic processes is to be designed and each equipment specifications are to be determined. A system runs continuously when it follows definite processes in a cyclic path. Process design means, determination of the type of thermodynamic processes included in the thermodynamic cycle and fixing the points i.e. pressure and temperature. While designing the process, equipment availability, constraints and cost should be kept in mind. Process design also includes the setting the parameters up to the optimum condition that maximum amount of liquid will be obtained.

Chapter 2

Literature Review

2. LITERATURE REVIEW

2.1 History of Liquefaction

Before 1877, a number of workers had discovered by visual observation in thick-walled glass tubes that the permanent gases, including hydrogen, nitrogen, oxygen and carbon monoxide, could not be liquefied at pressures as high as 400 atm. At first in 1877 oxygen gas is liquefied by Cailletet and Pictet. It is the first permanent gases to be liquefied. The term 'permanent' arose from the experimentally determined fact that such 'permanent' gases could not be liquefied by pressure alone at ambient temperature, in contrast to the non-permanent or condensable gases like chlorine, nitrous oxide and carbon dioxide, which could be liquefied at quite modest pressures of 30-50 atm.

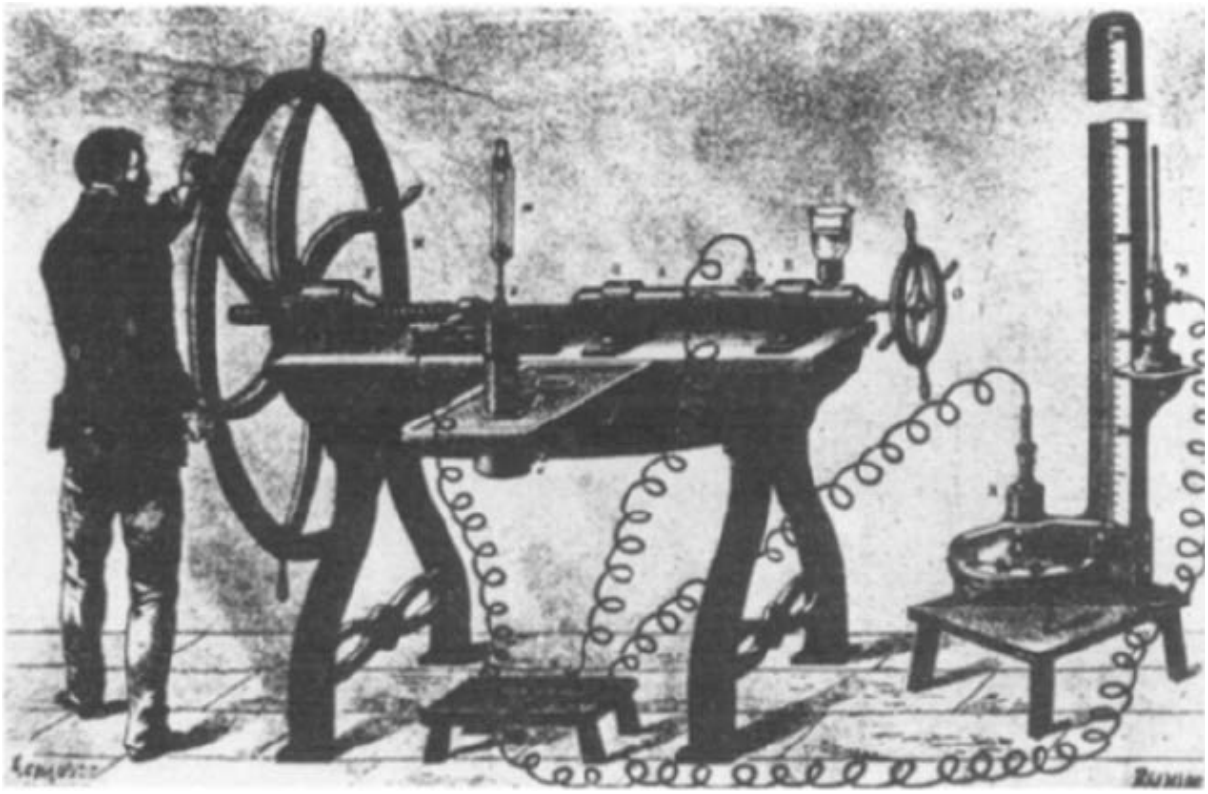

Fig. 2.1 Cailletet's gas compressor and liquefaction apparatus

Figures 2.1 show the apparatus which Cailletet used to produce a momentary fog of oxygen droplets in a thick walled glass tube. The oxygen gas was compressed using the crude Natterer compressor in which pressures up to 200 atm. were generated by a hand-operated screw jack. The pressure was transmitted to the oxygen gas in the glass tube by hydraulic transmission using water and mercury. The gas was cooled to -110°C by enclosing the glass tube with liquid ethylene, and was then expanded suddenly by releasing the pressure via the hand wheel. A momentary fog was seen, and the procedure could then be repeated for other observers to see the phenomenon.

Simultaneously at the first liquefaction of oxygen by Cailletet, Pictet also liquefied oxygen in the same year 1877.

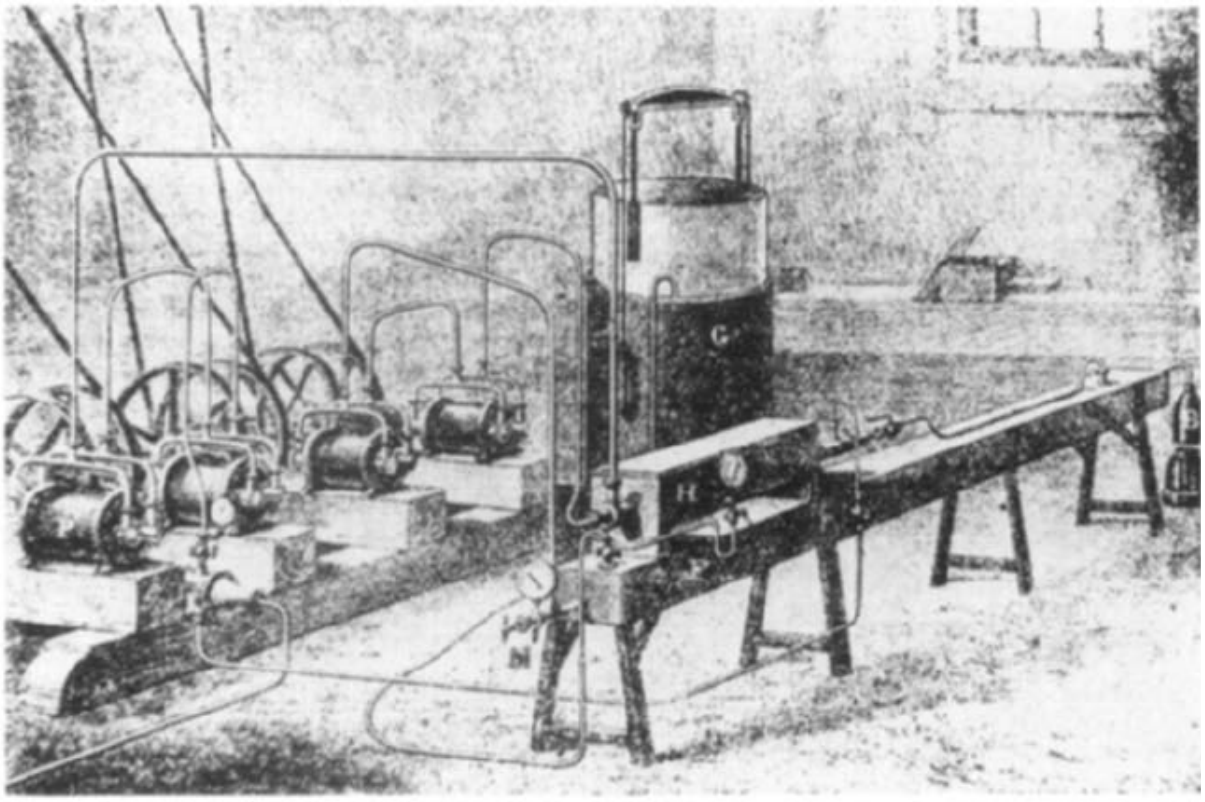

Fig 2.2 Pictet's cascade refrigeration and liquefaction system

Figure 2.2 shows the cascade refrigeration system of Pictet, in which oxygen was first cooled by sulphur dioxide and then by liquid carbon dioxide in heat exchangers, before being expanded into the atmosphere by opening a valve. The expansion yielded a transitory jet of liquid oxygen, but no liquid could be collected from the high velocity jet. The figure shows how Pictet used pairs of compressors to drive the SO2 (-20°C) and CO2 (-60°C) refrigerant cycles on a continuous basis, and this is probably the first example of a cascade refrigeration system operating at more than one temperature level. His use of the cascade system inspired others like Kamerlingh Onnes and Dewar.

In 1883, the Polish scientists Olzewski and Wroblewski, at Cracow, had improved Cailletet's apparatus by:

1. Adding an inverted U to the glass tube; and
2. Reducing the ethylene temperature to -136°C by pumping it below atmospheric pressure.

These improvements enabled them to produce small quantities of liquid oxygen in the U tube and to liquefy carbon monoxide and nitrogen for a few seconds.

From first liquefaction of oxygen to 1895, there was little progress in the developments of liquefiers. Then in 1895, Hampson in London and Linde in Munich simultaneously patented compact and efficient air liquefiers which used self-intensive or regenerative cooling of the high pressure air by the colder low pressure expanded air in long lengths of coiled heat exchanger. In this simple way, the complications of cascade precoolers employing liquid ethylene and other liquid cryogens were removed. A further advantage of this simple liquefier was the absence of moving parts at low temperature, the cooling being produced by Joule-Thomson expansion through a nozzle or valve. Carl von Linde made rapid progress in developing this technological breakthrough. He was a professor and research worker at the University of Munich, and he had his own company constructing refrigeration plant.

The Linde-Hampson is the simplest of all the liquefaction systems. A schematic of the Linde-Hampson system is shown in Fig. 2.3 and the cycle is shown on the T-s plane in Fig. 2.4.

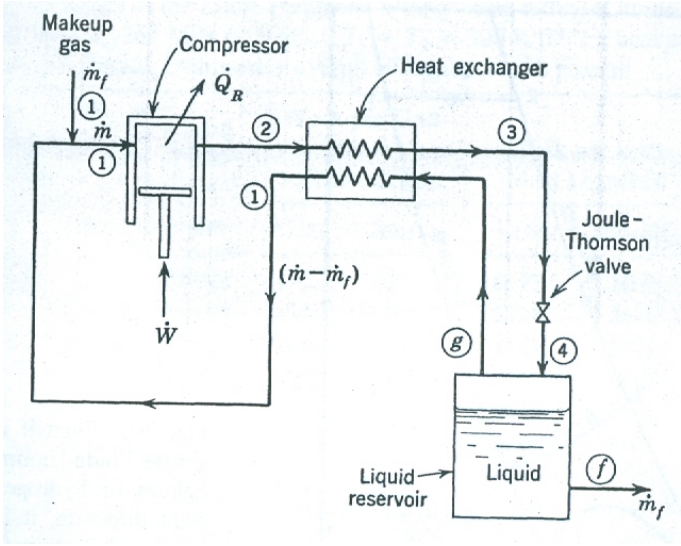

Fig. 2.3 Linde air liquefaction system

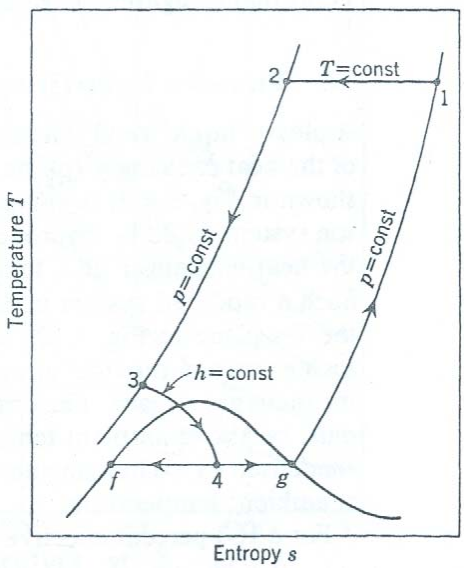

Fig. 2.4 T-S Diagram of Linde cycle

Process 1 to.2 would actually be two processes: an irreversible, adiabatic or polytropic compression followed by an after cooling to lower the gas temperature back to within a few degrees of ambient temperature. The gas next passes through a constant-pressure heat exchanger (ideally) in which it exchanges energy with the outgoing low pressure stream to point 3. From point 3 to point 4, the gas expands through an expansion valve to P4. At point 4, some of the gas stream is in the liquid state and is withdrawn at condition f (saturated-liquid condition), and the rest of the gas leaves the liquid receiver at condition g (saturated-vapor condition). This cold gas is finally warmed to the initial temperature by absorbing

energy at constant pressure (ideally) from the incoming high-pressure stream. The liquid air produced is very less.

By 1898, Charles Tripler, an engineer in New York, had constructed a similar but much larger air liquefier, driven by a 75 kW steam engine, which produced literally gallons of liquid air per hour. Tripler discovered a market for liquid air as a medium for driving air expansion engines (the internal combustion engine was still unreliable at that time) and succeeded to launch his Liquid Air Company.

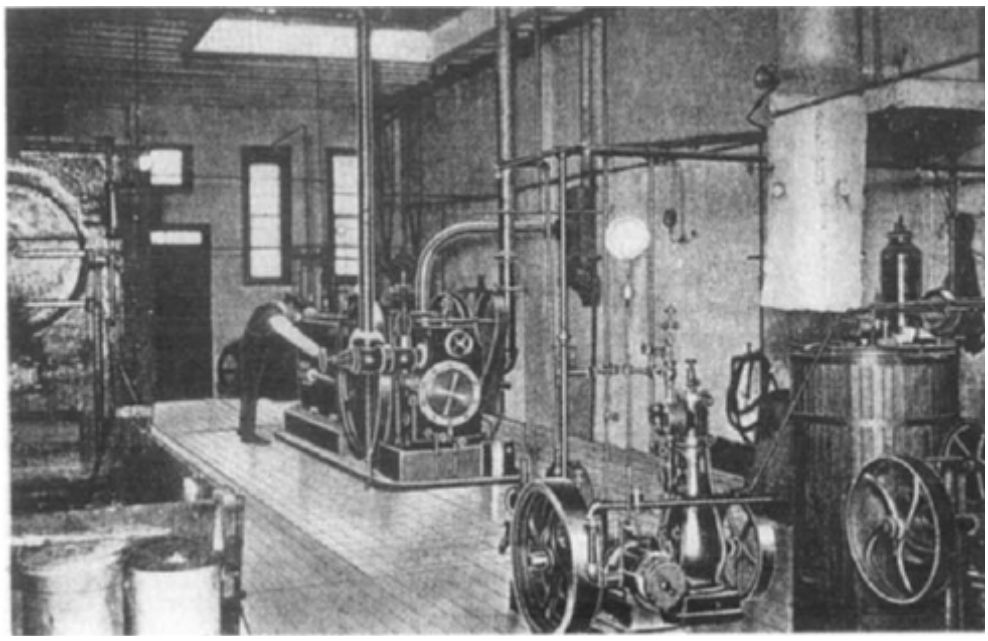

Fig. 2.5 Tripler's laboratory showing 175 kW steam driven multistage air compressor and 25 dm h- air liquefier

In the year, 1902, a young French innovative engineer Georges Claude, with wide connections in the scientific world of Paris, had succeeded in producing a piston expansion engine working at the low temperatures required for the liquefaction of air. The increase in cooling effect over the Joule-Thomson nozzle expansion of the Linde, Tripler, and Hampson designs was so large as to constitute a second technological breakthrough. Claude developed air liquefiers with piston expanders.

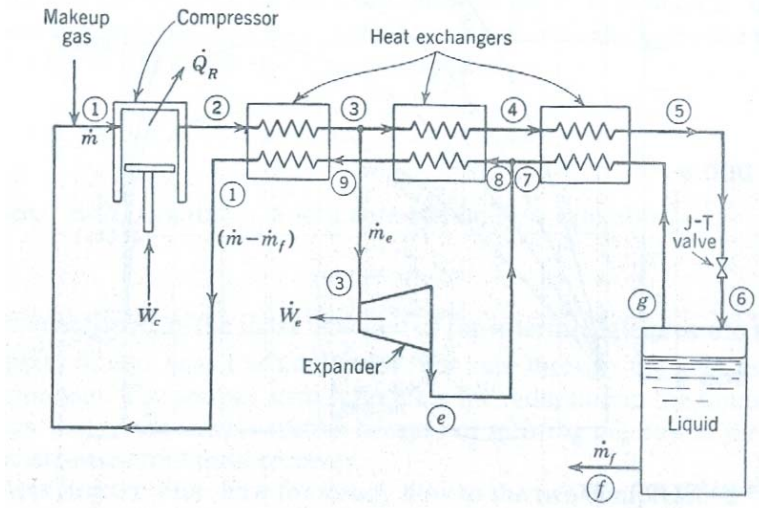

Fig. 2.6 Claude air liquefaction system

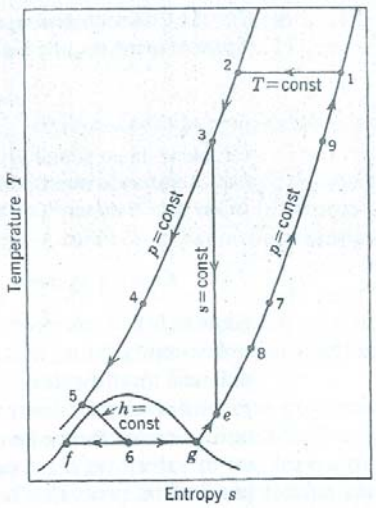

Fig. 2.7 T-S Diagram of Claude Cycle

The expansion through an expansion valve is an irreversible process. Thus if we wish to approach closer to the ideal performance, we must seek a better process to produce low temperatures. In the Claude system, shown in Fig. 2.6, energy is removed from the gas stream by allowing it to do some work in an expansion engine or expander.

The Claude cycle is shown on the T-s plane in Fig. 2.7. If the expansion engine is reversible and adiabatic, the expansion process is isentropic, and a much lower temperature is attained than for an isenthalpic expansion, In the Claude system, the gas is first compressed to pressures on the order of 4 MPa (40 atm or 590 psia) and then passed through the first heat exchanger. Between 60 and 80 percent of the gas is then diverted from the mainstream, expanded through an expander, and reunited with the return stream below the second heat exchanger. The stream to be liquefied continues through the second and third heat exchangers and is finally expanded through an expansion valve to the liquid receiver. The cold vapor from the liquid receiver is returned through the heat exchangers to cool the incoming gas.

In 1882, Kamerlingh Onnes set up a cryogenic laboratory at the University of Leiden in the Netherlands. In 1866, Van der Waals had published his first paper on 'the continuity of liquid and gaseous states' from which the physical understanding of the critical state and of liquefaction and evaporation was to grow. This information inspired Kamerlingh Onnes and for the first time in 1908, he was able to liquefy helium. He had only 360 liters of gaseous helium obtained by heating monazite sand from India. More than 60 cm^2 of liquid helium was produced by ones in his first attempt.

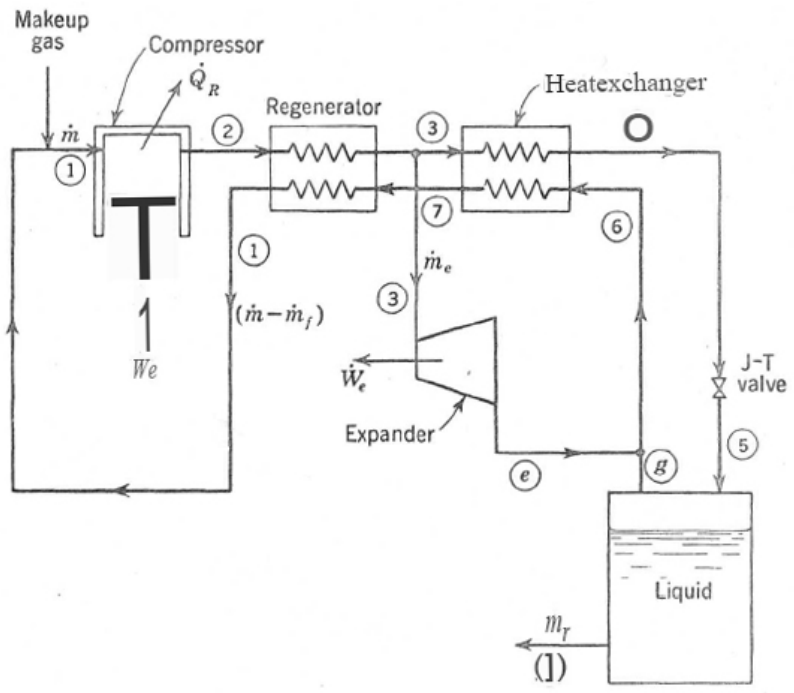

Fig. 2.8 Kapitza system

Kapitza (1939) modified the basic Claude system by eliminating the third or low temperature heat exchanger, as shown in Fig. 2.8. Several notable practical modifications were also introduced in this system. A rotary expansion engine was used instead of a reciprocating expander. The first or high-temperature heat exchanger in the Kapitza system was actually a set of valved regenerators, which combined the cooling process with the purification process. The incoming warm gas was cooled in one unit and impurities were deposited there, while the outgoing stream warmed up in the other unit and flushed out the frozen impurities deposited in it. After a few minutes, a valve was operated to cause the high- and low-pressure streams to switch from one unit to the other. The Kapitza system usually operated at relatively low pressures-on the order of 700 kPa (7 atm or 100 psia).

Around 1942 Samuel C. Collins of the department of mechanical Engineering at Massachusetts Institute of technology developed an efficient liquid helium laboratory facility. He developed Collins helium cryostat results economical and safe production of liquid helium.

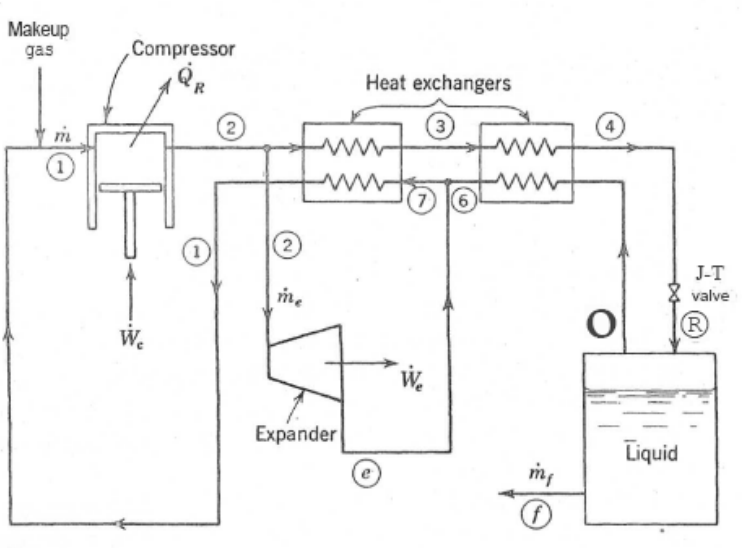

Fig. 2.9 Helandt system

Helandt (Davies 1949) noted that for a high pressure of approximately 20 MPa (200 atm) and an expansion-engine flow-rate ratio of approximately 0.60, the optimum value of temperature before expansion through the expander was near ambient temperature. Thus, one could eliminate the first heat exchanger in the Claude system by compressing the gas to 20 MPa. Such a modified Claude system is called the Heylandt system, after its originator, and is used extensively in high-pressure liquefaction plants for air. The system is shown schematically in Fig. 2.9. The advantage of the Heylandt system is that the lubrication problems in the expander are not difficult to overcome. In the air-liquefaction system, the gas enters the expander at ambient temperature and leaves the expander at approximately 150 K (-190°F), So that light lubricants can be used.

From time to time a lot of modifications had been made in all these cycles to optimize the results. The efficiency and performance of the components are increased, and so little modification in those cycles can reach up to lowest temperature and produce liquid for longer period.

Chapter 3

Process Design

3. PROCESS DESIGN

3.1 Modified Claude Cycle for Nitrogen Liquefier

A modified Claude cycle is taken into consideration to design nitrogen liquefier to take the advantage of both the turboexpander and JT valve. Instead of three heat exchangers as in the Claude cycle, two numbers of heat exchangers are used in this liquefier. Last two heat exchangers of the Claude cycle are combined to a single heat exchanger to reduce the cost of the liquefier.

A turbo expander based nitrogen liquefier consists of following parts:

- Compressor
- Heat exchangers
- Turboexpander
- JT Valve
- Phase separator
- Cold box
- Piping
- Instrumentation

A screw compressor will be installed to provide the compressed nitrogen gas. Heat exchangers are vital components of any cryogenic refrigerator. To exchange high heat in small area plate fin compact heat exchanger are used. The turboexpander is the heart of the liquefier and it can used lowering the temperature to expectable amount adiabatically. JT valve is used for isenthalpic expansion. Phase separator is used to separate liquid and gas. Piping and other instrumentations are required to connect and control the systems. Whole thing is kept inside the cold box.

Fig.3.1 shows the process diagram of the nitrogen liquefier. At atmospheric temperature and pressure at 1.1 bar the pure nitrogen gas is feed into the screw compressor and compressed up to 8 bars. The compressed gas is passed through the first heat exchanger i.e. HX1. Then some mass is diverted through the turboexpander and remain passes through the second heat exchanger i.e.HX2 for liquefaction. For easy calculation HX2 split into two parts i.e. HX2a and HX2b. From the HX2, isenthalpic expansion takes place by using JT valve which results liquid nitrogen. Liquid nitrogen taken out and remain vapor nitrogen meet with the isentropic expanded nitrogen by the turboexpander and feed again to the compressor by passing through the HX2 and HX1.

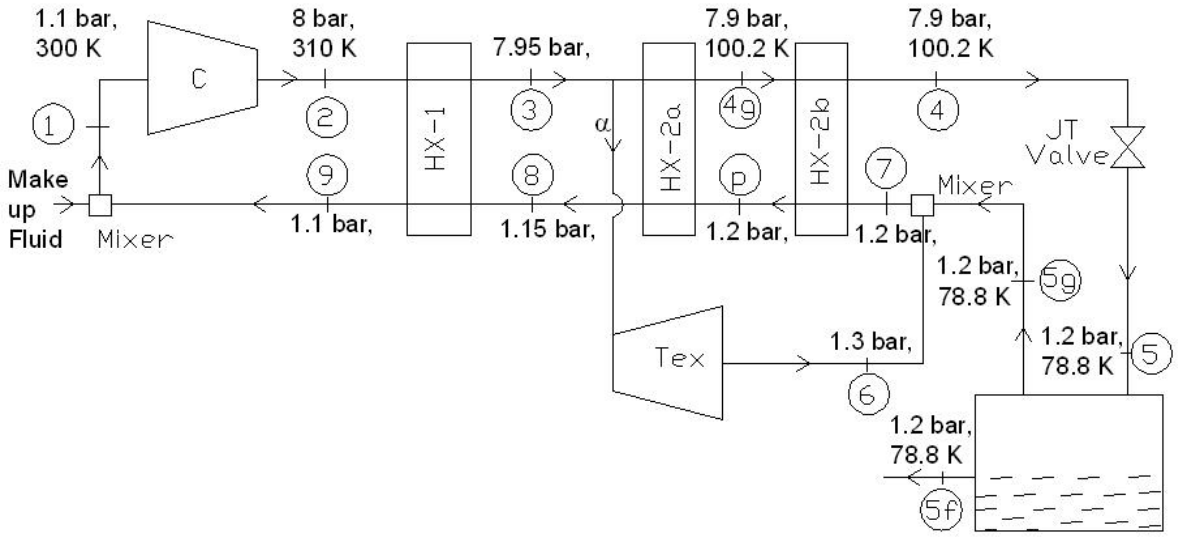

Fig. 3.1 Process Diagram Nitrogen Liquefier

3.2. Steps of the process design calculations

A. Known values

Pure nitrogen feed to the screw compressor at temperature, 300 K and pressure 1.1 bars. Isothermal compression is considered but in real case, temperature is increased. Let it is increased to the temperature 310 K and pressure is 8 bars.

Generally maximum pressure drop in both the heat exchangers is taken as 0.05 bars. Hence the pressure of high pressure stream after the HX1 is 7.95 bars. Similarly after HX2, the pressure is 7.9 bars. Passing through the HX2 Nitrogen in the high pressure stream comes to two phase state. The saturated temperature at 7.9 bars is 100.13 K.

The pressure inside the phase separator should be just higher than the atmospheric, so that the liquid will come out of the cold box. So phase separator pressure is fixed to 1.2 bars. Saturated temperature of nitrogen at 1.2 bars inside phase separator is 78.8 K.
Between HX2 and phase separator, JT Valve is placed for isenthalpic expansion of nitrogen from 7.9 bars to 1.2 bars.

The fraction of mass flow through the turboexpander expanded to the pressure 1.3 bars to maintain the pressure ratio of 6. So that mach No. should not exceed 1.

$$\frac{P_3}{P_6} = 6$$

Then from mixture comes out at 1.2 bars and return to compressor at 1.1 bars by pressure drop of 0.05 bars at each heat exchanger. Pressure difference is maintained at all equipments so flow will occur cyclically.

B. Parameters:

The parameters, by changing which, the amount of liquid nitrogen effected or by controlling which we optimize the output are:

- Effectiveness of heat exchanger 1,$\varepsilon 1$
- Pinch point for heat exchanger 2, p
- Efficiency of turbo expander,η
- Mass flow ratio diverted through Turbo expander, α

C. Unknown Variables:

Following are the unknown variables which value are to be determined.

- Enthalpy at the exit of first heat exchanger, h_3
- Enthalpy at the exit of second heat exchanger, h_4
- Enthalpy at the exit of JT valve, h_5
- Enthalpy at the exit of turboexpander, h_6
- Enthalpy at the exit of turboexpander (Isentropic expansion), h_{6s}
- Enthalpy at after mixing from phase separator and exit of turboexpander, h_7
- Enthalpy at the inlet to the second heat exchanger, h_8
- Enthalpy at the inlet to the first heat exchanger, h_9
- Dryness fraction in the phase separator, x_5
- Ratio of mass of liquid produced to mass of gas compressed, yield, y
- Enthalpy at the pinch point temperature of low pressure stream of second heat exchanger, h_p

D. Component Analysis

Some initial value of Yield, y and low pressure stream outlet enthalpy from HX1, $h_9 = \varepsilon 1 h_{2'}$ should be taken.

i. Pinch point specification of Heat exchanger-2

Splitting the HX2 into two parts, First heat exchanger being the one where the hot nitrogen gas is cooled up to the saturation temperature of 100.13 K & the second part being

the condensing part. The minimum temperature difference occurs at the point where the condensation begins and is called as pinch point.

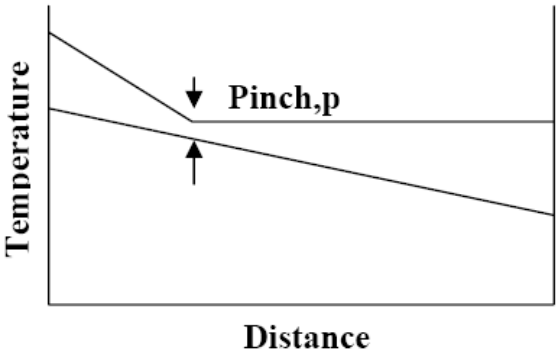

Fig. 3.2 Pinch point of HX2

For the specified pinch value p, for HX2, we have

$$T_{4g} - T_p = p,$$

$$T_p = T_{4g} - p \tag{3.1}$$

We can get enthalpy h_p, at that pinch temperature and pressure.

ii. Heat Exchanger-1

For the specified value of effectiveness of heat exchanger 1 and the pinch point specification for HX2, h_8, h_3 and h_9 calculated from the effectiveness definition and energy balance between hot and cold fluids for HX1 and HX2a.

Assume

$$h_8 = \frac{[h_9(1-y)(1-\alpha) - h_2(1-\alpha) + h_{4g}(1-\alpha) - h_p(1-y)]}{[(-\alpha)(1-y)]} \tag{3.2}$$

$$h_3 = \frac{[h_{4g}(1-\alpha) + (1-y)(h_8 - h_p)]}{(1-\alpha)} \tag{3.3}$$

$$h_9 = \varepsilon 1 h_{2'} + (1-\varepsilon 1) h_8 \tag{3.4}$$

The updated value of h_9 is calculated. It should be checked with the previous value. This iteration should be done until both are equal.

iii. Turbo-expander

From the Fig. 3.2, it is clear that 3-6s is the isentropic expansion and 3-6 is the actual expansion.

From property table, entropy at 3, i.e. s_3, can be found out at h_3 and p_3.

The enthalpy at the end of expansion is found out as

$$s_{6s} = s_3 \tag{3.5}$$

h6s can be get from p6s and s6s.

$$h_6 = h_3 - \eta(h_3 - h_{6s}) \quad (3.6) \tag{6}$$

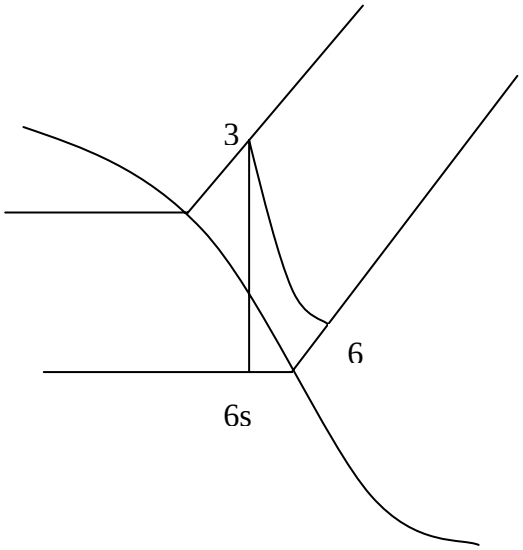

Fig. 3.3 Expansion in Turbo Expander

vi. Mixer

Applying energy balance equation for the mixer, enthalpy at outlet of mixer is

$$(1-\alpha-y)h_{5g} + \alpha h_6 = (1-y)h_7$$

$$h_7 = \frac{[\alpha h_6 + (1-\alpha-y)h_{5g}]}{(1-y)} \tag{3.7}$$

v. Heat Exchanger 2

Enthalpy at outlet of hot fluid is found out by energy balance between hot and cold fluids as

$$h_4 = \frac{[h_3(1-\alpha) - (1-y)(h_8 - h_7)]}{(1-\alpha)} \tag{3.8}$$

vi. Throttle valve:

Throttling is an isenthalpic process. Equating the enthalpies before and after throttling

$$h_5 = h_4 \tag{3.9}$$

$$x_5 = \frac{(h_5 - h_{f5})}{(h_{g5} - h_{f5})} \qquad (3.10)$$

vii. Yield:

The liquid yield obtained per kg of gas passing through the throttling valve is therefore $(1\text{-}x_5)$.

For $(1-\alpha)$ kg of gas passing through the throttling valve is

$$y = (1-\alpha)(1-x_5) \qquad (3.11)$$

Again check the calculated value of y with the assumed one and calculated h_9 with assumed h_9. Replace it with the new value of h_9 and y and calculate again till both the assumed values match with calculated value

3.3 Process Design Calculation Using Microsoft Excel

Thermodynamic properties like pressure, temperature, enthalpy, entropy all are calculated by using the above equations and following the above procedure. Below there is table showing all the thermodynamic property values and liquid yield produced.

Working Fluid	:	Nitrogen				
Asumed Yeild , Y	=	0.043262				
Flow Through Turbo Expander,α	=	0.94				
Effectiveness of HX1	=	0.98				
Efficiency of Expander	=	0.5				
Pinch temperature HX2	=	1				
Mass Flow Rate	=	296	kg/hr	=	0.08222	kg/s
Assumed Enthalpy at 9	=	317.461				

Points	Presssure (bar)	Temperature (K)	Enthalpy kJ/kg	Entropy kJ/kg/K	Quality	Mass Flow Rate kg/s
1	1.1	300.00	311.44	6.8194		0.08222
2	8	310.00	320.44	6.2602		0.08222
2*	1.1	310.00	321.85	6.8535		
3	7.95	120.52	114.76	5.2269		0.07728
4	7.9	100.13	-64.24	3.4653	0.0541	0.00493
4f	7.9	100.13	-72.91	3.3788	0.0000	
4g	7.9	100.13	87.39	4.9786	1.0000	
5	1.2	78.78	-64.24	3.5678	0.2790	0.00493
5f	1.2	78.78	-119.19	2.8708	0.0000	0.00355
5g	1.2	78.78	77.78	5.3693	1.0000	0.00137
6s	1.3	79.49	68.24	5.2269	0.9487	
6sf	1.3	79.49	-117.72	2.8892	0.0000	
6sg	1.3	79.49	78.29	5.3533	1.0000	
6	1.3	90.81	91.496	5.5090	> 1	0.07728
7	1.2	90.40	91.26	5.5293		0.07866
P	1.15	99.13	100.77	5.6420		
8	1.15	100.73	102.48	5.6592		0.07866
9	1.1	305.78	317.461	6.8393		0.07866

Calculated Yeild , Y in Phase Separator	=	0.043262	
Energy Balance	=	0.043261	
Temperature Drop in Turbo Expander	=	29.71	
Cumulative UA for Hx 1	=	1.8787	kW/K
Cumulative UA for Hx 2	=	0.2032	kW/K
Liquid Nitrogen Produced		12.805206	kg/hr

* Note :Text written inside double lined border are the input values

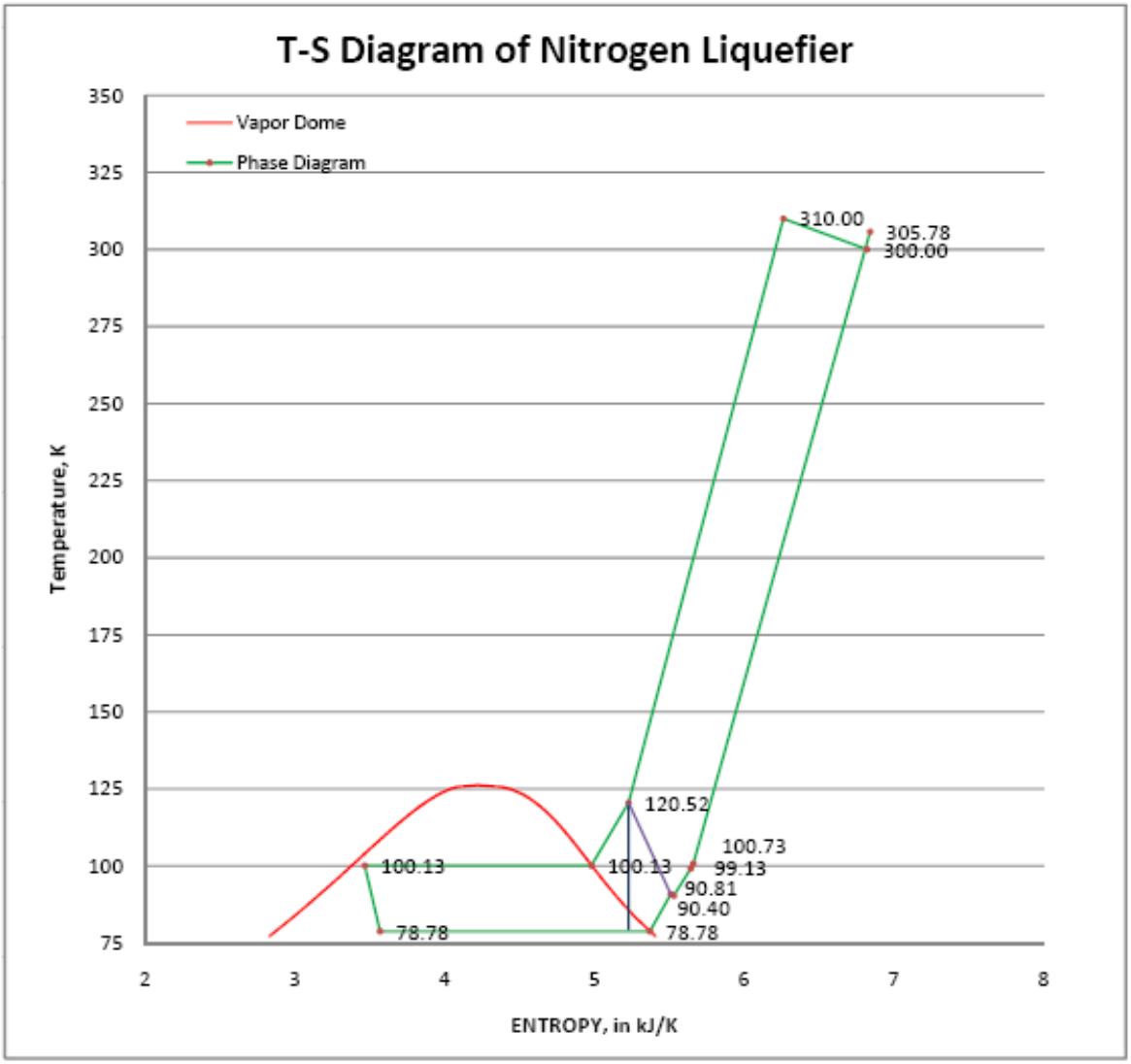

Fig. 3.4 T-S Diagram of Nitrogen Liquefier

3.4 Process Design Using Aspen Hysys

3.4.1 Introduction to Aspen Hysys

Aspen Hysys is a process simulation environment designed to serve many processing industries especially Oil & Gas and Refining. With Aspen Hysys one can create rigorous steady state and dynamic models for plant design, performance monitoring, troubleshooting, operational improvement, business planning, and asset management. Through the completely interactive Aspen Hysys interface, one can easily manipulate process variables and unit operation topology, as well as fully customize your simulation using its customization and extensibility capabilities. The process simulation capabilities of Aspen Hysys enables engineers to predict the behavior of a process using basic engineering relationships such as mass and energy balances, phase and chemical equilibrium, and reaction kinetics. With reliable thermodynamic data, realistic operating conditions and the rigorous Aspen Hysys equipment models, they can simulate actual plant behavior. Some of the important Aspen Hysys features are listed below:

- Windows® Interoperability **:** Interface contains a process flow sheet view for graphical layout, data browser view for entering data, the patented Next expert guidance system to guide the user through a complete and consistent definition of the process flow sheet.

- Plot Wizard: Hysys enables the user to easily create plots of simulation results.

- Flowsheet Hierarchy and Templates: Collaborative engineering is supported through hierarchy blocks that allow sub-flowsheets of greater detail to be encapsulated in a single high-level block. These hierarchy blocks can be saved as flowsheet templates in libraries.

- Equation-Oriented Modeling: Advanced specification management for equation oriented model configuration and sensitivity analysis of the whole simulation or specific parts of it. The unique combination of Sequential Modular and Equation Oriented solution technology allows the user to simulate highly nested processes encountered typically in the chemical industry.

- Thermo physical Properties: Physical property models and data are key to generating accurate simulation results that can be used with confidence. Aspen Hysys uses the extensive and proven physical property models, data and estimation methods available in Aspen Properties™, which covers a wide range of processes from simple ideal behavior to strongly non-ideal mixtures and electrolytes. The built-in database contains parameters for more than 8,500 components, covering organic, inorganic, aqueous, and salt species and more than 37,000 sets of binary interaction parameters for 4,000 binary mixtures.

- Convergence Analysis: to automatically analyze and suggest optimal tear streams, flowsheet convergence method and solution sequence for even the largest flowsheets with multiple stream and information recycles.

- Sensitivity Analysis: to conveniently generate tables and plots showing how process performance varies with changes to selected equipment specifications and operating conditions.

- Design Specification: capabilities to automatically calculate operating conditions or equipment parameters to meet specified performance targets.

- Data-Fit: to fit process model to actual plant data and ensure an accurate, validated representation of the actual plant.

- Determine Plant Operating Conditions that will maximize any objective function specified, including process yields, energy usage, stream purities and process economics.

- Simulation Basic Manager: This feature available in Aspen Hysys for using different fluids like nitrogen, air, acetylene as per requirement. Also several fluid packages like BWRS, MWRS, and ASME are provided to calculate properties at different states.

3.4.2 Procedure of Process Design in Aspen Hysys

To create a new case, From the File menu, select New. In the sub-menu, select Case. The Simulation Basis Manager window will appear.

The Simulation Basis Manager is the main property view of the Simulation environment. One of the important concepts that HYSYS is based upon is Environments. The Simulation Basis environment allows you to input or access information within the Simulation Basis manager while the other areas of HYSYS are put on hold avoiding unnecessary Flowsheet calculations. Once you enter the Simulation environment, all changes that were made in the Simulation Basis environment will take effect at the same time. Conversely, all thermodynamic data is fixed and will not be changed as manipulations to the Flowsheet take place in the Simulation environment. The minimum information required before leaving the Simulation Basis manager is atleast one installed Fluid Package with an attached Property Package and At least one component in the Fluid Package.

The Components Manager is located on the Components tab of the Simulation Basis Manager. This tab provides a location where sets of chemical components being modeled may be retrieved and manipulated. These component sets are stored in the form of Component Lists that may be a collection of library pure components or hypothetical components.The Components Manager always contains a Master Component List that cannot be deleted. This master list contains every component available from "all" component lists. If you add components to any other component list, they automatically get added to the Master Component List. Also, if you delete a component from the master, it also gets deleted from any other component list that is using that component.

In HYSYS, all necessary information pertaining to pure component flash and physical property calculations is contained within the Fluid Package. This approach allows you to define all the required information inside a single entity. There are four key advantages to this approach:

- All associated information is defined in a single location, allowing for easy creation and modification of the information.
- Fluid Packages can be exported and imported as completely defined packages for use in any simulation.

- Fluid Packages can be cloned, which simplifies the task of making small changes to a complex Fluid Package.
- Multiple Fluid Packages can be used in the same simulation.

The Fluid Package Manager is located on the Fluid Pkgs tab of the Simulation Basis Manager. This tab provides a location where multiple fluid packages can be created and manipulated. Each fluid package available to your simulation is listed in the Current Fluid packages group with the following information: name, number of components attached to the fluid package, and property package attached to the fluid package. From the Fluid Pkgs tab of the Simulation Basis Manager click either the View or Add button to open the Fluid Package property view. Make sure you select the proper fluid package when using the view option. Click on the Set Up tab. From the Component List Selection drop-down list, select the components you want to use in your fluid package.

Here Benedict-Webb-Rubin-Starling (BWRS) fluid package was used. This model is commonly used for compression applications and studies. It is specifically used for gas phase components that handle the complex thermodynamics that occur during compression, and is useful in both upstream and downstream industries.

After selecting fluid packages and components , a process flowsheet window will apear on which the unit opearations can be installed. There are a number of ways to install unit operations into your flowsheet. Many unit operations are available in the flowsheet palette. All information concerning a unit operation can be found on the tabs and pages of its property view. Each tab in the property view contains pages, which pertain to a certain aspect of the operation, such as its stream connections, physical parameters (for example, pressure drop and energy input), or dynamic parameters such as vessel rating and valve information. In steady state analysis recycler unit operations can be used to calculate the unknown parameters in the process flow diagram.

The process flow diagram (PFD) provides the best representation of the flowsheet as a whole. Using the PFD gives you immediate reference to the progress of the simulation currently being built, such as what streams and operations are installed, flowsheet connectivity, and the status of objects. In addition to graphical representation, you can build your flowsheet within the PFD using the mouse to install and connect objects. A full set of manipulation tools is available so you can reposition streams and operations, resize icons, or reroute streams. All of these tools are designed to simplify the development of a clear and

concise graphical process representation. The PFD also possesses analytical capabilities. You can access property views for streams or operations directly from the PFD, or install custom Material Balance Tables for any or all objects. Complete Workbook pages can also be displayed on the PFD and information is automatically updated when changes are made to the process.

3.4.3 Input values in hysys

From simulation basis manager in the component pure nitrogen is taken as material stream and BWRS as fluid packages. Then enter into the simulation environment. There all unit operations are arranged in order and linked by material streams. For each unit operations follwing input values are entered.

1. Compressor

Mass flow rate =296 kg/hr

Inlet temperature = 300 K

Inlet pressure = 1.1 bar

Outlet pressure =8 bar

2. Cooler

Outlet temperature = 310 K

Outlet pressure =8 bar

3. Heat exchanger 1

Minimum Approach = 4.15 K

Pressure drop in both streams =0.05 bar

4. Tee

Flow ratio through turbo expander = 0.94

5. Turbo expander

Efficiency of turbo expander = 50 %

Outlet pressure =1.3 bar

6. Heat exchanger 2

Minimum Approach = 1 K

Pressure drop in both streams =0.05 bar

7. JT Valve

Outlet pressure = 1.2 bar

3.4.4 Results in Aspen Hysys

Amount of liquid yeild can be seen in the liquid stream of the phase separator. It comes 12.558 kg/hr.

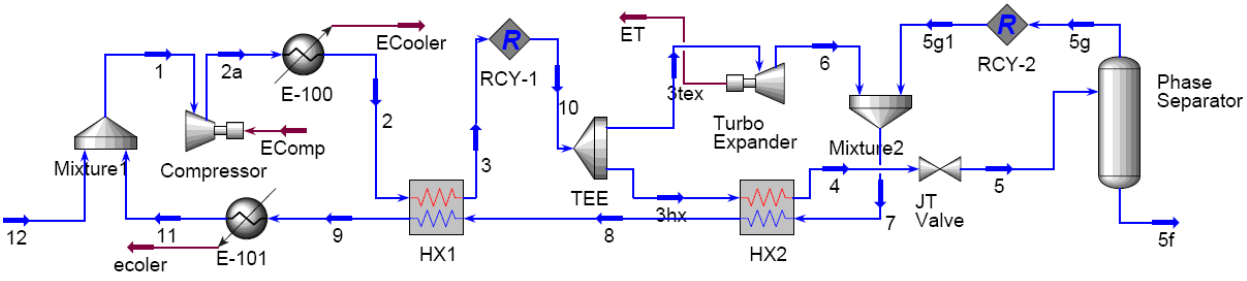

Fig. 3.5 Process Flow Diagram of Nitrogen Liquefier in Aspen Hysys

Figure 3.5 shows the process flow diagram that drawn in Hysys. Table 3.1 shows the state and properties of all the streams in the process flow diagram mass flow at 5_f, gives the liquid nitrogen that produced.

Table 3.1 Material stream properties in Aspen Hysys

Material Streams							
		1	**2a**	**2**	**8**	**3tex**	**3hx**
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	K	300.00	595.78	310.00	100.98	120.94	120.94
Pressure	bar	1.100	8.000	8.000	1.150	7.950	7.950
Molar Flow	kgmole/h	10.57	10.57	10.57	10.12	9.933	0.6340
Mass Flow	kg/h	296.00	296.00	296.00	283.45	278.24	17.760
Liquid Volume Flow	m3/h	0.3671	0.3671	0.3671	0.3515	0.3451	2.202e-002
Heat Flow	kW	0.1381	26.12	0.8789	-16.00	-14.79	-0.9442
		4	**7**	**5**	**5g**	**5f**	**6**
Vapour Fraction		0.0086	1.0000	0.2929	1.0000	0.0000	1.0000
Temperature	K	100.35	90.032	79.161	79.161	79.161	90.441
Pressure	bar	7.900	1.200	1.200	1.200	1.200	1.300
Molar Flow	kgmole/h	0.6340	10.12	0.6340	0.1857	0.4483	9.933
Mass Flow	kg/h	17.760	283.45	17.760	5.2020	**12.558**	278.24
Liquid Volume Flow	m3/h	2.202e-002	0.3515	2.202e-002	6.451e-003	1.557e-002	0.3451
Heat Flow	kW	-1.855	-16.92	-1.855	-0.3274	-1.528	-16.59

		5g1	3	10	9	11	12
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	K	79.160	120.94	120.94	305.85	300.00	300.00
Pressure	bar	1.200	7.950	7.950	1.100	1.100	1.100
Molar Flow	kgmole/h	0.1858	10.57	10.57	10.12	10.12	0.4482
Mass Flow	kg/h	5.2056	296.00	296.00	283.45	283.45	12.554
Liquid Volume Flow	m3/h	6.456e-003	0.3671	0.3671	0.3515	0.3515	1.557e-002
Heat Flow	kW	-0.3276	-15.74	-15.74	0.6125	0.1322	5.857e-003

Chapter 4

Results and Discussion

4. RESULTS AND DISCUSSION

4.1 Performance Analysis

The effect of parametric variation is done on the liquefaction system gives the optimum performance. This analysis also depicts the off design performance analysis. The parameters are

- Effectiveness of heat exchanger 1,ε1
- Pinch point for heat exchanger 2, p
- Efficiency of turbo expander,η_t
- Mass flow ratio diverted through Turbo expander, α

4.1.1 Effect of Variation of expander flow ratio, α

The value of turbine efficiency, effectiveness of HX1, and pinch point of HX2 are kept constant. The effect of yield with the variation of mass flow ratio through turboexpander is studied as shown in figure 4.1. Yield increases with the increase in mass fraction through turboexpander. But after a mass fraction liquid does not produced. This indicates that there is an optimum yield at a particular mass flow ratio through the turboexpander.

Table 4.1 Effect of variation of expander flow ratio

Turbo expander flow ratio ,α	Yield (Excel)	Yield (Hysys)	Compressor work per kg of liquid produced in kJ/kg (Excel)	Compressor work per kg of liquid produced in kJ/kg (Hysys)
0.94	0.04278	0.04211	210.36	213.75
0.90	0.04096	0.04030	219.71	223.32
0.85	0.03870	0.03804	232.58	236.59
0.80	0.03644	0.03585	246.98	251.08
0.70	0.03197	0.03128	281.55	287.77
0.60	0.02756	0.02704	326.52	332.86
0.50	0.02327	0.02249	386.76	400.16
0.40	0.01917	0.01827	469.53	492.70
0.30	0.01543	0.01459	583.39	616.91
0.20	0.01251	0.01091	719.25	824.59

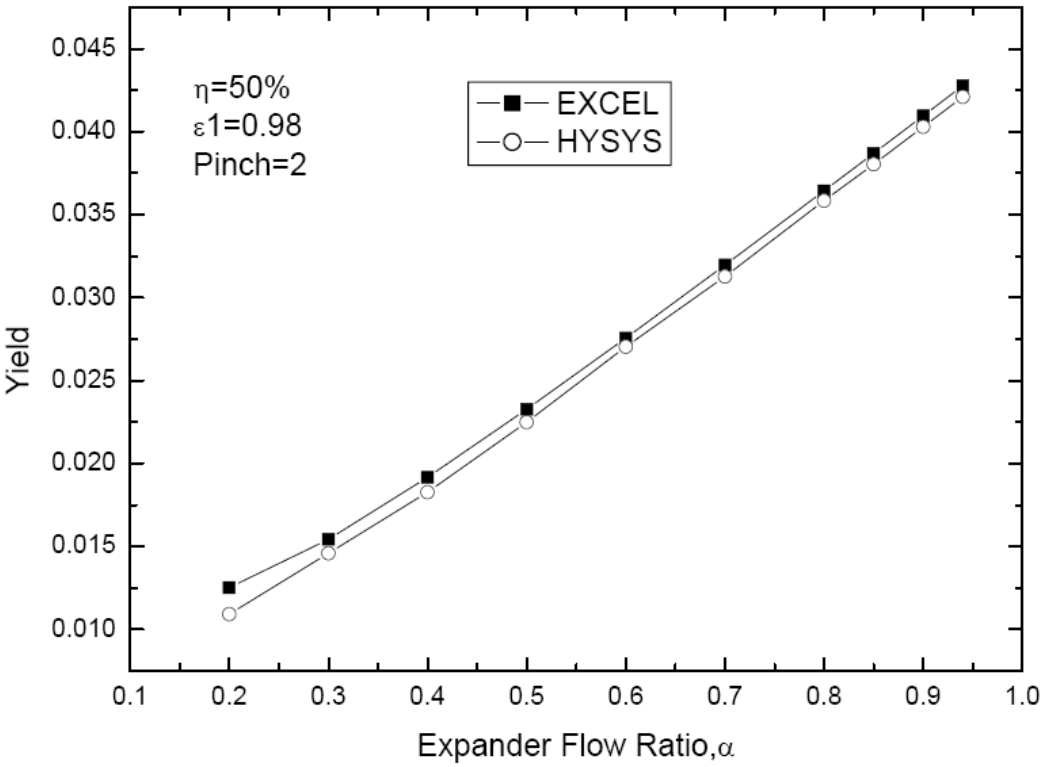

Fig. 4.1 Variation of yield with expander flow ratio

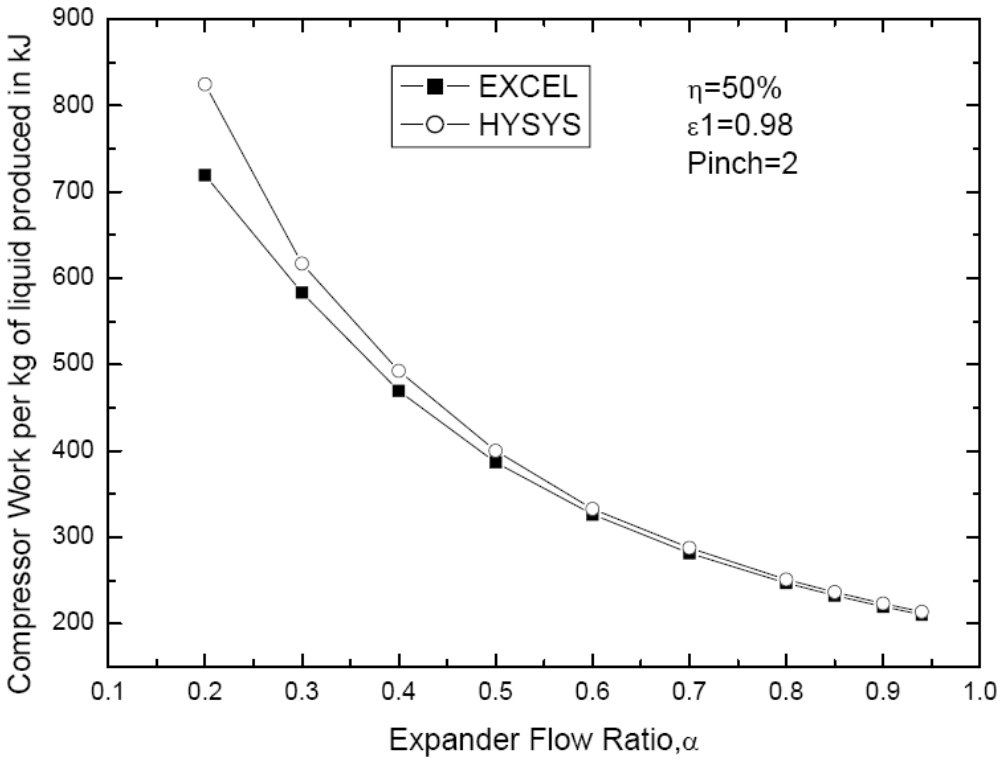

Fig. 4.2 Variation of compressor work per kg of liquid with expander flow ratio

4.1.2 Effect of Variation of effectiveness of HX1, ε1

The value of turbine efficiency, mass fraction flow through turboexpander, and pinch point of heat exchanger-2 are kept constant. The effect of yield with the variation of effectiveness of heat exchanger-1 is studied as shown in figure. There is a lower limitation of effectiveness of HX1 below which liquid does not come. Figure shows that yield increases with increase in effectiveness of HX1. But there is manufacturing limitations of effectiveness.

Table 4.2 Effect of Variation of effectiveness HX1

Effectiveness of Heat exchanger 1,ε1	Yield (Excel)	Yield (Hysys)
0.98	0.04278	0.04211
0.97	0.03838	0.03784
0.96	0.03395	0.03365
0.95	0.02948	0.02937
0.94	0.02499	0.02503
0.92	0.01591	0.01628
0.90	0.00671	0.00756
0.89	0.00206	0.00304

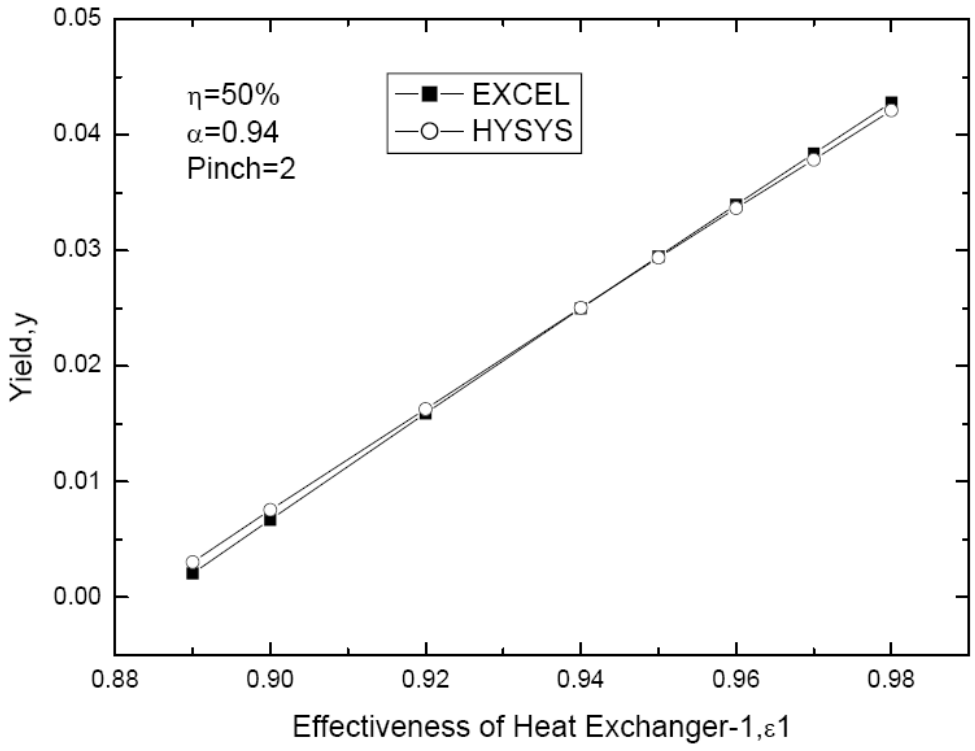

Fig. 4.3 Variation yield with effectiveness of HX-1

4.1.3 Effect of Variation of pinch point of second heat exchanger

The value of turbine efficiency, mass fraction flow through turboexpander, and effectiveness of first heat exchanger are kept constant. The effect of yield with the variation of pinch point of second heat exchanger is studied as shown in figure. Pinch point does not affect much on yield but decreasing the pinch temperature performance of HX2 will be improved.

Table 4.3 Effect of Variation of pinch point of HX-2

Pinch Temperature Of heat exchanger 2	Yield (Excel)	Yield (Hysys)
1.00	0.04326	0.04258
2.00	0.04278	0.04211
3.00	0.04231	0.04162
4.00	0.04185	0.04123
8.00	0.04005	0.03944
12.00	0.03836	0.03783

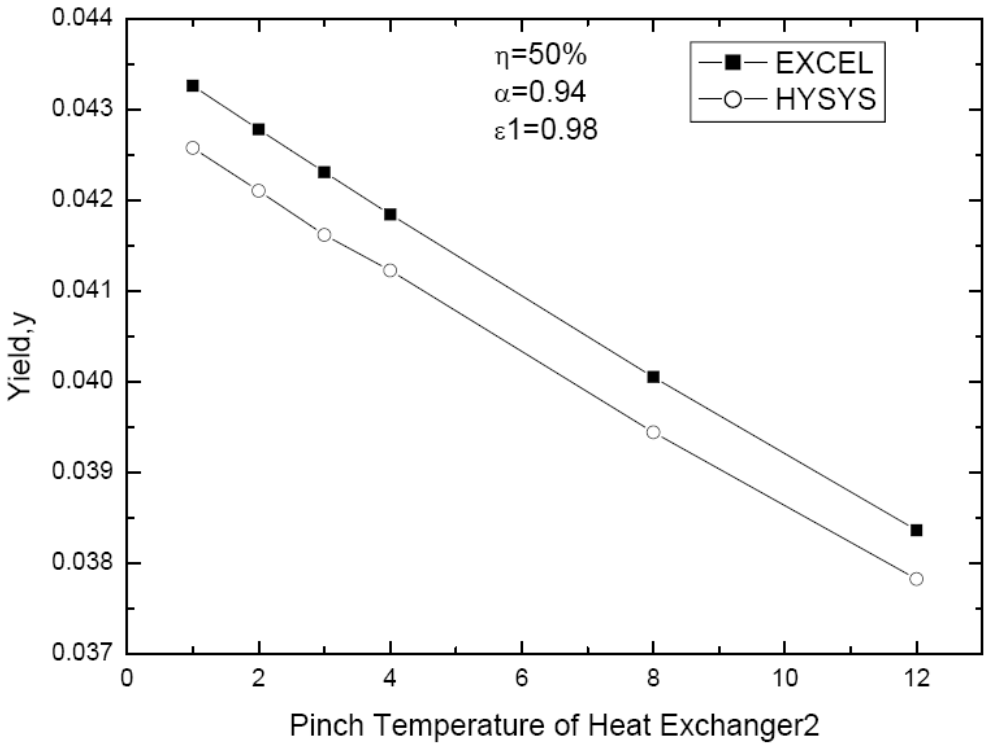

Fig. 4.4 Variation yield with Pinch temperature of HX-2

4.1.4 Effect of Variation of turbo expander efficiency η

The value of mass ratio through turboexpander, effectiveness of first heat exchanger and pinch temperature of HX2 are kept constant. The effect of yield with the variation of of turbo expander efficiency is studied as shown in figure. Yield increases with the increase in the efficiency of turboexpander. But due to availability of the turboexpander, it is limited to 50%.

Table 4.4 Effect of Variation of turbo expander efficiency

Turbo expander efficiency, η	Yield (Excel)	Yield (Hysys)
0.40	0.03215	0.03172
0.50	0.04278	0.04211
0.60	0.05385	0.05186

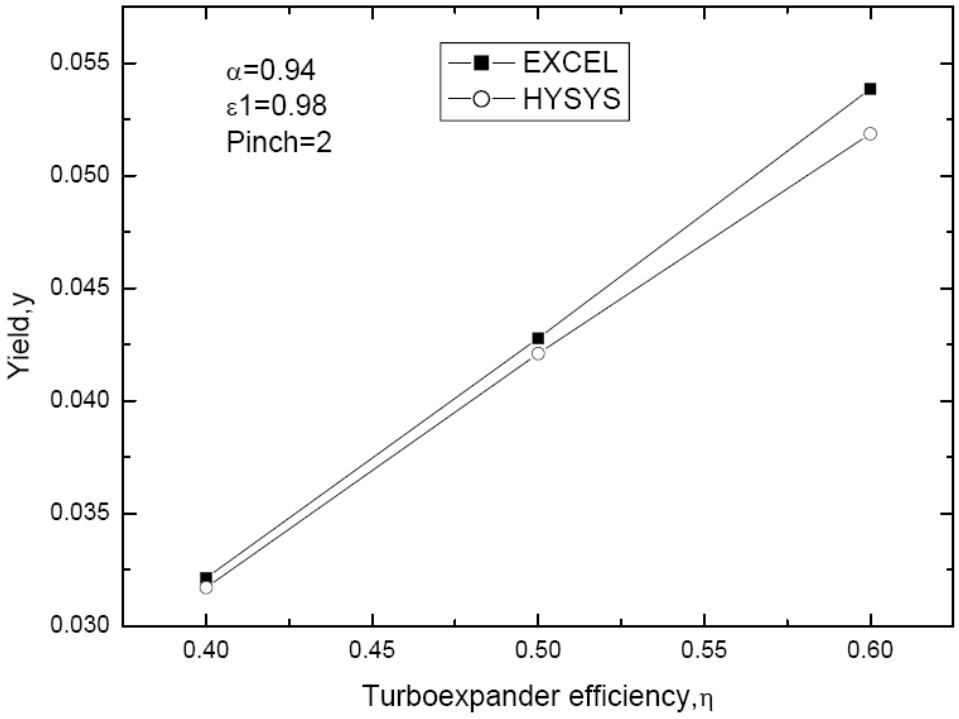

Fig. 4.5 Variation yield with Turboexpander efficiency

4.2 Variable Specific Heat Analysis of Heat Exchangers

NTU-effectiveness relationships, was integrated with the restriction that the specific heats of the fluids was constant. The fluid properties do vary considerably in the near critical region, and cryogenic heat exchangers may operate in this regime. The specific heat also varies significantly in a condenser in which the fluid enters as a superheated vapor.Chowdhry and Sarangi (1984b) examined the effect of variable specific heats on the performance of hydrogen heat exchangers. Oonk and Hustevedt (1986) examined the same effect for helium heat exchangers. Soyars (1991) examined the effect of variable fluid properties on the accuracy of analysis of helium heat exchanger performance in the temperature range below 15 K .Their results indicated that noticeable errors result for helium heat exchangers in refrigeration systems if the helium specific heat was treated as a constant below 15 K.

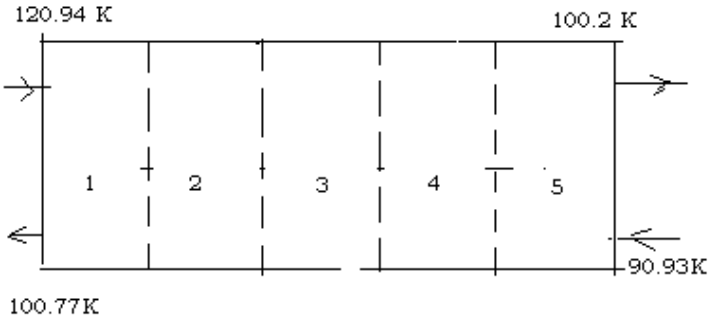

Fig.4.6 Heat exchanger split into parts

A technique similar to the finite element approach will be used; the heat exchanger will be subdivided into small elements, in which the specific heat variation is relatively small and may be treated as constant. In the heat exchanger 2, Temperature of the hot fluid was divided into five parts and their enthalpies found out by using Allprops. The number of Transfer units for each element is found out as follows:

Enthalpy of cold fluid at inlet of element 1 is found out from energy balance between hot and cold fluids as

$$m_h(i_{h1} - i_{h2}) = m_c(i_{c2} - i_{c1})$$

$$i_{c1} = \frac{[i_{c2}m_c + m_h(i_{h2} - i_{h1})]}{m_c} \quad (1)$$

Heat capacities of cold and hot streams are found as

$$C_h = \frac{(i_{h1} - i_{h2})}{(T_{h1} - T_{h2})} \quad (2)$$

$$C_c = \frac{(i_{c2} - i_{c1})}{(T_{c2} - T_{c1})} \quad (3)$$

$$C_R = \frac{C_{\min}}{C_{\max}}$$

The effectiveness of Heat exchanger 1 is found out as

$$\varepsilon = \frac{(T_{c2} - T_{c1})}{(T_{h1} - T_{c1})} \qquad if \quad C_c < C_h \tag{4}$$

Or else,

$$\varepsilon = \frac{(T_{h1} - T_{h2})}{(T_{h1} - T_{c1})} \tag{5}$$

The number of transfer units for each element is found out as

$$Ntu = \frac{\log_e\{(1 - C_R\varepsilon)/(1-\varepsilon)\}}{(1 - C_R)} \tag{6}$$

$$UA = Ntu \times C_{\min} \tag{7}$$

4.2.1 Analysis of Heat exchanger-1

Heat exchanger-1 is divided into ten elements. Heat transfer area required is calculated considering the specific heat variation. Fig 4.7 shows the operating temperature line of HX1. The distance of the operating temperature line from equilibrium line gives the minimum temperature difference or pinch point temperature. In HX1 temperature pinch occurs at hot end of the heat exchanger.

Table 4.5: Calculation of UA of HX1 considering variable specific heat

Element No. from hot end	Temp. in Cold stream (K)	Enthalpy in Cold stream (kJ/kg)	Enthalpy in Hot stream (kJ/kg)	Temp. in Hot stream (K)	Heat Capacity of cold stream (kJ)	Heat Capacity of cold stream (kJ)	Heat capacity ratio (Cr)	Effectiv-eness	Ntu	UA in kW/K
0	305.78	317.45	320.44	310.00						
1	285.28	296.11	300.02	290.59	0.0819	0.0865	0.9468	0.8294	4.3239	0.3541
2	264.77	274.76	279.60	271.21	0.0819	0.0867	0.9446	0.7943	3.4992	0.2866
3	244.27	253.40	259.15	251.88	0.082	0.0869	0.9436	0.7611	2.9299	0.2402
4	223.76	232.00	238.68	232.60	0.0821	0.0873	0.9404	0.7294	2.4997	0.2052
5	203.26	210.57	218.18	213.40	0.0822	0.0878	0.9362	0.6989	2.1646	0.1779
6	182.75	189.10	197.64	194.31	0.0824	0.0885	0.9311	0.6691	1.8931	0.156
7	162.25	167.58	177.05	175.37	0.0826	0.0894	0.9239	0.6396	1.6647	0.1375
8	141.74	145.99	156.39	156.66	0.0828	0.0908	0.9119	0.6098	1.4642	0.1212
9	121.24	124.31	135.65	138.32	0.0832	0.093	0.8946	0.5789	1.2838	0.1068
10	100.73	102.48	114.77	120.59	0.0837	0.0969	0.8638	0.5456	1.1121	0.0931
								Cum. UA =		1.8787
								Avg Cc =		0.0825
								Avg Ch =		0.0893
									NTU=	22.782

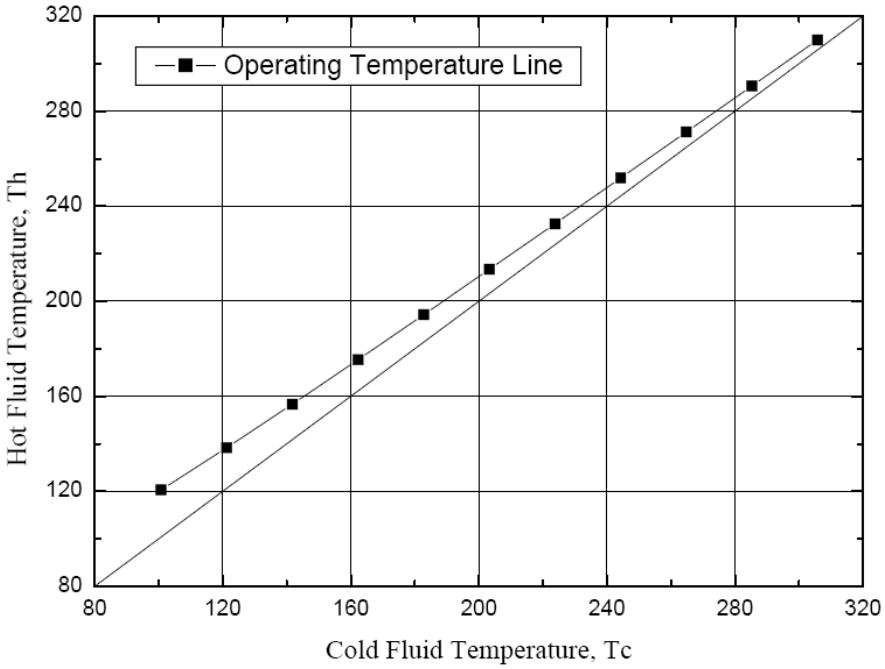

Fig. 4.7 Operating Temperature Line for Heat exchanger 1

4.2.2 Analysis of Heat exchanger-2

HX2 is divided into six elements. From element 0 to 5 are in single phase while 5 to 6 is in two phase. Fig 4.8 shows the operating temperature line of HX2. The distance of the operating temperature line from equilibrium line gives the minimum temperature difference or pinch point temperature. In HX2 temperature pinch occurs at the time of change of phase.

Table 4.6: Calculation of UA of HX2 considering variable specific heat

Element No. from hot end	Temp in Hot stream (Th)	Enthalpy in Hot stream (hh)	Enthalpy in Cold stream (hc)	Temp in Cold stream (Tc)	Heat Capacity of cold stream (Cc)	Heat Capacity of hot stream (Ch)	Heat capacity ratio (Cr)	Effectiveness	Ntu	UA in (kW/K)
0	120.52	114.76	102.48	100.73						
1	116.44	109.75	102.17	100.44	0.0842	1.1327	0.0743	0.0145	0.0146	0.0012
2	112.36	104.61	101.85	100.14	0.0842	1.1620	0.0725	0.0183	0.0185	0.0016
3	108.28	99.28	101.52	99.83	0.0843	1.2057	0.0699	0.0246	0.0249	0.0021
4	104.20	93.61	101.16	99.50	0.0843	1.2816	0.0658	0.0383	0.0391	0.0033
5	100.13	87.39	100.77	99.13	0.0843	1.4106	0.0598	0.0718	0.0747	0.0063
6	100.13	-64.24	91.26	90.31	0.0848		0.0000	0.8985	2.2872	0.1940
								Cum. UA=		0.2084

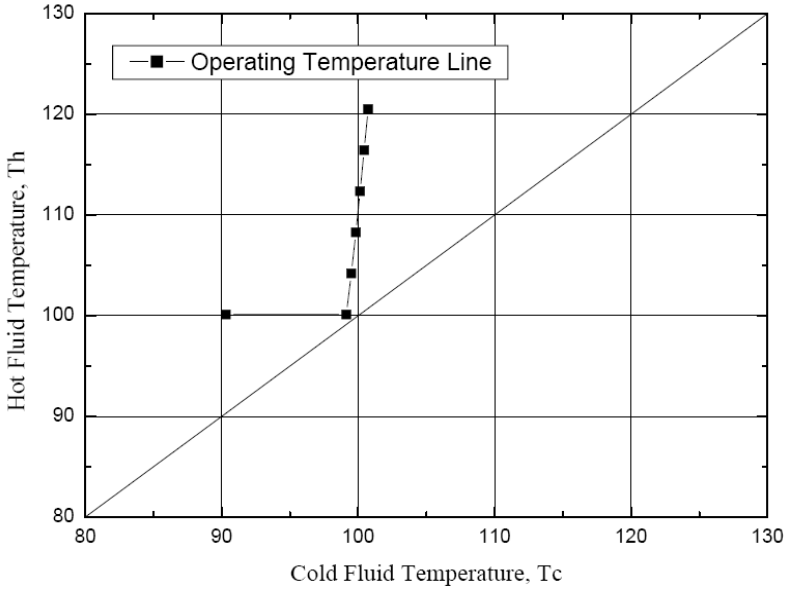

Fig. 4.8 Operating Temperature Line for Heat exchanger 2

4.3 Cumulative Enthalpy Analysis of Heat Exchangers

Cumulative enthalpy analysis is done to find the variation of temperature along the length of the heat exchanger. It will also shows the whether there is any temperature cross.

4.3.1 Analysis of Heat exchanger-1

Table 4.7 Cumulative enthalpy in L.P. Side of HX-1

Temp. (k)	L P side (1.1 bar) Cp	Change in Temp	Average Cp	Mass Flow Rate	Change in Enthalpy	Cum Enthalpy
305.78	1.0404	0.00	1.04036	0.07866	0.000	0.000
280	1.0409	25.78	1.0406	0.07866	2.110	2.110
260	1.0417	20.00	1.0413	0.07866	1.638	3.748
240	1.0428	20.00	1.0423	0.07866	1.640	5.388
220	1.0443	20.00	1.0436	0.07866	1.642	7.030
200	1.0461	20.00	1.0452	0.07866	1.644	8.674
190	1.0471	10.00	1.0466	0.07866	0.823	9.497
180	1.0483	10.00	1.0477	0.07866	0.824	10.321
170	1.0496	10.00	1.0489	0.07866	0.825	11.146
160	1.0510	10.00	1.0503	0.07866	0.826	11.972
150	1.0526	10.00	1.0518	0.07866	0.827	12.799
145	1.0535	5.00	1.0531	0.07866	0.414	13.213
140	1.0545	5.00	1.0540	0.07866	0.415	13.628
135	1.0556	5.00	1.0551	0.07866	0.415	14.043
130	1.0568	5.00	1.0562	0.07866	0.415	14.458
125	1.0582	5.00	1.0575	0.07866	0.416	14.874
120	1.0597	5.00	1.0590	0.07866	0.416	15.290
115	1.0615	5.00	1.0606	0.07866	0.417	15.707
110	1.0636	5.00	1.0626	0.07866	0.418	16.125
105	1.0662	5.00	1.0649	0.07866	0.419	16.544
100.73	1.0690	4.27	1.0676	0.07866	0.369	**16.913**

Table 4.8 Cumulative enthalpy in H.P. Side of HX-1

Temp. (k)	H P side (8 bar) Cp	Change in Temp	Average Cp	Mass Flow Rate	Change in Enthalpy	Cum Enthalpy
310.00	1.0507	0.00	1.0507	0.08222	0.000	0.000
300.00	1.0516	10.00	1.0512	0.08222	0.864	0.864
280.00	1.0541	20.00	1.0529	0.08222	1.731	2.595
260.00	1.0577	20.00	1.0559	0.08222	1.736	4.331
240.00	1.0624	20.00	1.0600	0.08222	1.743	6.074
220.00	1.0688	20.00	1.0656	0.08222	1.752	7.826
200.00	1.0778	20.00	1.0733	0.08222	1.765	9.591
190.00	1.0836	10.00	1.0807	0.08222	0.889	10.480
180.00	1.0907	10.00	1.0871	0.08222	0.894	11.374
170.00	1.0995	10.00	1.0951	0.08222	0.900	12.274
160.00	1.1107	10.00	1.1051	0.08222	0.909	13.183
150.00	1.1254	10.00	1.1181	0.08222	0.919	14.102
145.00	1.1345	5.00	1.1300	0.08222	0.465	14.567
140.00	1.1453	5.00	1.1399	0.08222	0.469	15.036
135.00	1.1582	5.00	1.1518	0.08222	0.474	15.510
130.00	1.1740	5.00	1.1661	0.08222	0.479	15.989
125.00	1.1937	5.00	1.1838	0.08222	0.487	16.476
120.52	1.2163	4.48	1.2050	0.08222	0.444	**16.920**

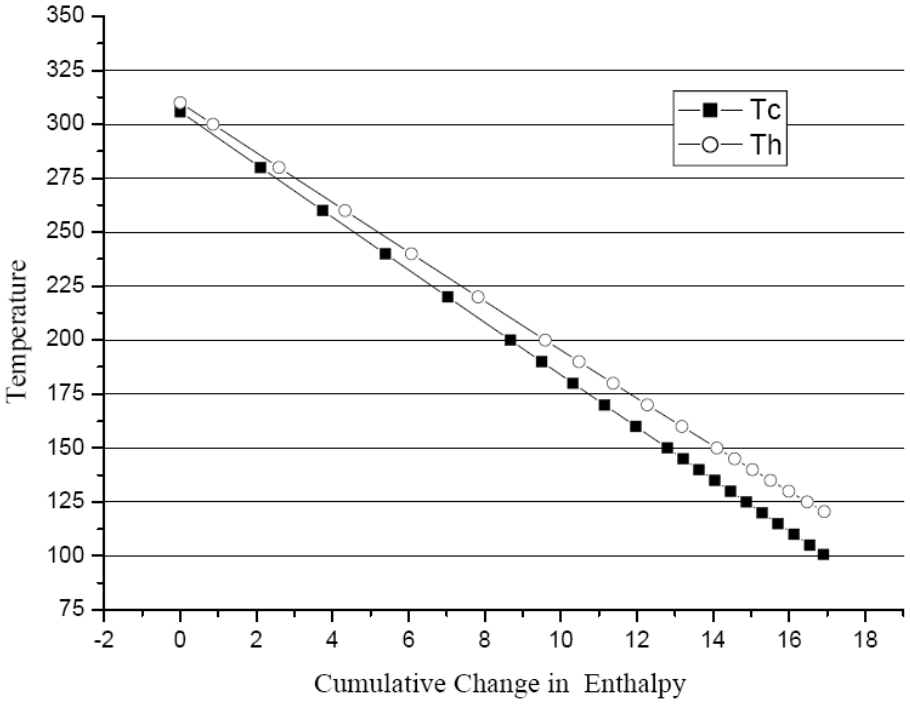

Fig. 4.9 Cumulative enthalpy- Temp Diagram for Heat Exchanger 1

4.3.2 Analysis of Heat exchanger-2

Table 4.9 Cumulative enthalpy in L.P. Side of HX-2

Temp. (K)	L P side 1.15 bar Cp	Change in Temp	Average Cp	Mass Flow Rate	Change in Enthalpy	Cum Enthalpy
100.73	1.0705	0.00	1.0705	0.0787	0.000	0.000
99.00	1.0720	1.73	1.0713	0.0787	0.146	0.146
98.00	1.0730	1.00	1.0725	0.0787	0.084	0.230
97.00	1.0741	1.00	1.0735	0.0787	0.084	0.315
95.00	1.0766	2.00	1.0754	0.0787	0.169	0.484
93.00	1.0800	2.00	1.0783	0.0787	0.170	0.653
90.40	1.0865	2.60	1.0833	0.0787	0.221	0.874

Table 4.10 Cumulative enthalpy in H.P. Side of HX-2

Temp. (K)	H P side 7.95 bar Cp	Change in Temp	Average Cp	Mass Flow Rate	Change in Enthalpy	Cum Enthalpy
120.52	1.0425	0.00	1.0425	0.00493	0.000	0.000
118.00	1.0423	2.52	1.0424	0.00493	0.013	0.013
116.00	1.0422	2.00	1.0422	0.00493	0.010	0.023
112.00	1.0419	4.00	1.0420	0.00493	0.021	0.044
110.00	1.0417	2.00	1.0418	0.00493	0.010	0.054
108.00	1.0415	2.00	1.0416	0.00493	0.010	0.064
106.00	1.0413	2.00	1.0414	0.00493	0.010	0.075
104.00	1.0412	2.00	1.0412	0.00493	0.010	0.085
101.00	1.0408	3.00	1.0410	0.00493	0.015	0.100
100.13	1.0407	0.87	1.0408	0.00493	0.004	0.105
100.13	PHASE CHANGE			0.00493	0.748	0.873

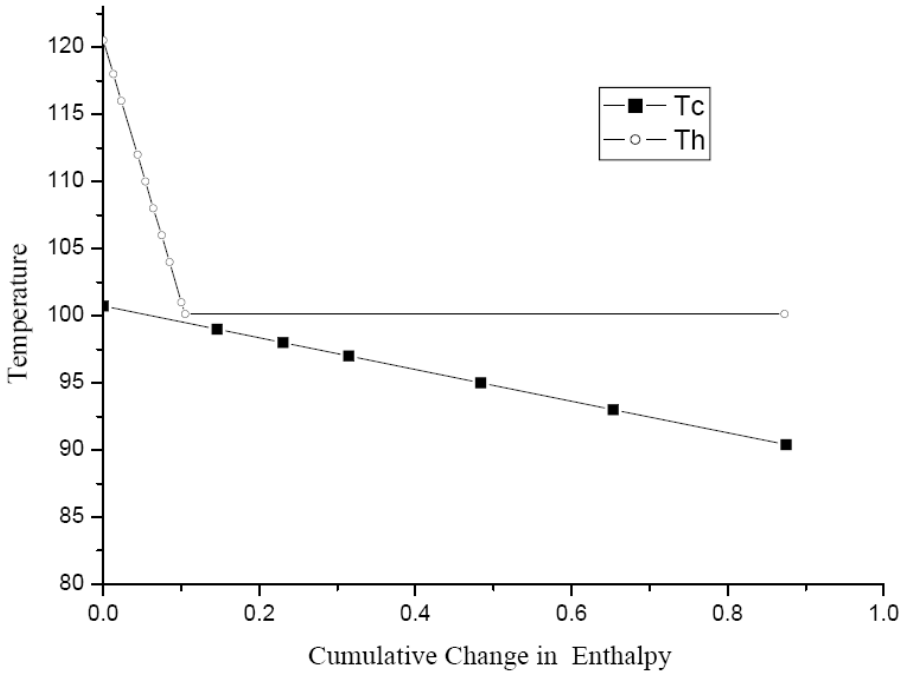

Fig. 4.10 Cumulative enthalpy- Temp Diagram for Heat Exchanger 2

Chapter 5

Conclusions

CONCLUSIONS

The process design is carried out using the standard calculation procedure and is validated by using process simulation software, Aspen Hysys. The preliminary data required in terms of mass flow rate, pressure and temperatures across heat exchanger, turbo expander and other components of the nitrogen liquefier are found out. Parametric study is carried out to study the role of the different component efficiencies in deciding overall system efficiency. It is found that the liquid yield is directly proportional to the effectiveness of heat exchanger, efficiency of turbo expander and mass fraction diverted through turbo expander. However, on the limit, no yield condition prevails if effectiveness of HX-1 is less than 0.88 and mass fraction through the turbo expander exceeds 0.95.The effect of increase in pinch point deteriorates the yield. Though higher turbine efficiency favors the yield, the turbine available for the purpose is limited to 50% efficiency and it limits the yield to only about 4%. The simulation done and the analysis carried out can serve as guide lines for the development of nitrogen liquefier in our nation and for the helium liquefier in particular as a future mission.

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